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Engineering and Economic Assessment of Autoheated Thermophilic Aerobic Digestion with Air Aeration

Camp, Dresser and McKee, Inc.
Boston, MA

Prepared for

Municipal Environmental Research Lab.
Cincinnati, OH

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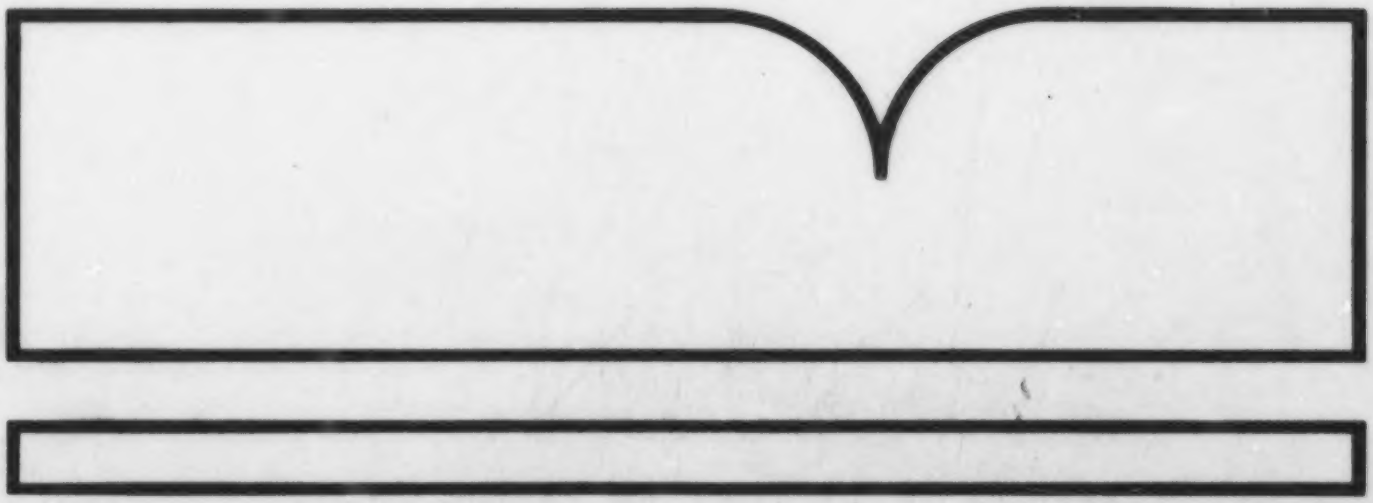

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ENGINEERING AND ECONOMIC ASSESSMENT OF AUTOHEATED THERMOPHILIC AEROBIC DIGESTION WITH AIR AERATION

by

Camp Dresser & McKee Inc.
Boston, MA 02108

Contract No. 68-03-2803

Project Officer

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16. ABSTRACT

A major disadvantage of aerobic digestion is that it requires long detention times, particularly in colder climates, to insure adequate stabilization. Autoheated thermophilic aerobic digestion (ATAD) offers the potential to decrease the required detention time. ATAD takes advantage of the heat generated by microorganisms as they degrade organic material. Heat losses are controlled so that the system will be self heating (authothermal) and reach thermophilic range (at least 45 ^{o}C). This study performs engineering and economic analyses of earlier work done on ATAD. ATAD has been shown to be feasible with a mixture of thickened waste activated and primary municipal sludge. ATAD systems appear economical compared with ambient aerobic digestion and anaerobic digestion. The system is also more effective in decreasing the pathogen content of sludge. However, further testing is needed to investigate the proper selection of an aerator. These studies should provide designers with adequate design tools so that the system capabilities could be maximized.

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FOREWORD

The Environmental Protection Agency was created because of increasing public and government concern about the dangers of pollution to the health and welfare of the American people. Noxious air, foul water, and spoiled land are tragic testimony to the deterioration of our natural environment. The complexity of that environment and the interplay between its components require a concentrated and integrated attack on the problem.

Research and development is that necessary first step in problem solution and it involves defining the problem, measuring its impact, and searching for solutions. The Municipal Environmental Research Laboratory develops new and improved technology and systems for the prevention, treatment, and management of wastewater and solid and hazardous waste pollutant discharges from municipal and community sources, for the preservation and treatment of public drinking water supplies, and to minimize the adverse economic, social, health, and aesthetic effects of pollution. This publication is one of the products of that research; a most vital communications link between the researcher and the user community.

In autothermal thermophilic aerobic digestion of sludge (ATAD), advantage is taken of the heat generated by the microorganisms as they degrade organic material. Controlling the heat losses allows the system to be self heating (autothermal) and to reach the thermophilic range. In conventional aerobic digestion, long detention times must be provided for in winter. With ATAD, the detention times can be greatly diminished. With this advantage, ATAD can be less expensive than anaerobic digestion and conventional (or ambient) aerobic digestion. ATAD merits further study and implementation.

Francis T. Mayo, Director
Municipal Environmental Research
Laboratory

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ABSTRACT

A major disadvantage of aerobic digestion is that it requires long detention times, particularly in colder climates, to insure adequate stabilization. Autoheated thermophilic aerobic digestion (ATAD) offers the potential to decrease the required detention time. ATAD takes advantage of the heat generated by microorganisms as they degrade organic material. Heat losses are controlled so that the system will be self heating (autothermal) and reach thermophilic range (at least 45^{o}C).

This study performs engineering and economic analyses of earlier work done on ATAD. ATAD has been shown to be feasible with a mixture of thickened waste activated and primary municipal sludge. ATAD systems appear economical compared with ambient aerobic digestion and anaerobic digestion. The system is also more effective in decreasing the pathogen content of sludge.

However, further testing is needed to investigate the proper selection of an aerator. These studies should provide designers with adequate design tools so that the systems capabilities could be maximized.

This report was submitted in fulfillment of Contract No. 68-03-2803 Work Effort No. 7 by Camp, Dresser, and McKee under the sponsorship of the U. S. Environmental Protection Agency. This report covers the period January 1980 to July 1980. Work was completed as of February 1981.

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This report was prepared by Camp Dresser and McKee Inc. under the supervision of Paul W. Prendiville and the direction of Albert B. Pincince. Project Engineer was Paul W. Keohan, assisted by Patrick J. Connelly.

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SECTION 1

INTRODUCTION

Aerobic digestion is a simple unit process for stabilizing sludge. A major disadvantage, particularly in colder climates, is that aerobic digestion requires long detention times to insure adequate stabilization.

To overcome this problem researchers have been investigating the use of autoheated thermophilic aerobic digestion (ATAD) to decrease the detention time, and hence the volume requirements. In ATAD, advantage is taken of the heat generated by the microorganisms as they degrade organic material. The intent is to control the heat losses so that the system will be self-heating (autothermal) and reach the thermophilic range (at least 45°C).

Andrews and Kambhu (1970) developed a steady-state mathematical model investigating the parameters affecting the process. Although much of the data on sludge characteristics and heat losses had to be assumed, these studies supported the basic theory that ATAD can be self-sustaining with respect to temperature. They suggested that pilot testing of ATAD be undertaken with both air and high-purity oxygen in order to test the general validity of the mathematical model.

Matsch and Drnevich (1977) investigated high-purity oxygen for aerating aerobic digesters. They found that aerobic digesters using oxygen can maintain elevated temperatures, and that these temperatures can reach the thermophilic range if the sludge contains a sufficient concentration of biodegradable solids. Their tests were conducted both at full-scale facilities at Speedway, Indiana and on bench-scale operations. Matsch and Drnevich did not compare their results with mathematical models.

Jewell et al presented the first successful large-scale application of an autoheated thermophilic aerobic digestion (ATAD) process with air aeration on a typical municipal sludge. (Others had either used oxygen to achieve thermophilic temperatures, had conducted their studies on a very small scale, or had not reached thermophilic temperatures.)

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Jewell et al. showed that ATAD with air aeration is feasible. They enumerated practical operating problems, developed relationships between operating variables, defined aeration requirements, measured the effect of high temperature on dewaterability and determined the effectiveness of ATAD on the destruction of pathogens.

This process is presented in the report entitled "Autoheated Aerobic Thermophilic Digestion with Air Aeration" (EPA Project No. R804636), and summarized in the March 1980 edition of the Journal of Water Pollution Control Federation (pp. 512-523). The report presented here analyzes the ATAD work done by Jewell et al. and performs an economic analysis of it.

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SECTION 2

CONCLUSIONS

We believe that autoheated thermophilic aerobic digestion (ATAD) with air aeration is feasible on a thickened municipal sludge.

Reaction rate constants calculated from the tests by Jewell et al were in the same range as those in previous studies, except in one experimental condition. In that case, the calculated constant greatly exceeded previous estimates.

Although ATAD appears promising, it is not without some problems. The self-aspirating aerators used by Jewell et al were unable to completely mix the digester contents. Moreover, those aerators might be a cause for the poor dewaterability of the digested sludge. Both of these deficiencies might be solved by the use of submerged-turbine aerators or other systems with efficient oxygen transfer characteristics.

Comparative analyses show that ATAD systems appear less expensive than ambient aerobic digestion and mesophilic anaerobic digestion. The cost estimates were developed for 3,800, 38,000 and 380,000-m^3/d (1,10 and 100-mgd) wastewater treatment plants using conservative criteria.

Jewell et al reported better pathogen reductions in their tests, than in the full-scale anaerobic digesters at the Binghamton-Johnson City plant, where the test program was conducted. In the test by Jewell et al, sludge addition was stopped for 12 to 24 hours before sludge was sampled for pathogens. This operation is typical for small plants. Pathogen reduction would be less in an operation where sludge is fed continuously. Continuous operation would be more typical in large plants.

The scope of work called for recommendations regarding the effects of heat exchangers and series operation of the digesters. Both of these modifications would decrease the size of the units required. Data are not presently available, however, to determine how much the digesters could be decreased. And, the economic viability of these changes cannot now be assessed without further mathematical modeling or full-scale testing.

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SECTION 3

RECOMMENDATIONS

There are several areas that need further attention, in order to maximize the advantages of ATAD. These areas include further development of mathematical models, aeration and mixing, digesters in series, heat exchangers, use of pure oxygen, and scale up.

MATHEMATICAL MODELS

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Evaluation of ATAD system would be greatly aided if the mathematical model developed by Andrews and Kambhu (1970) were extended. That work had shown trends regarding the influence of variables such as sludge concentration, volatile solids content, and retention time. But the report by Andrews and Kambhu (1970) did not examine all the possible combinations of variables.

An extension of that work would be valuable to assist the designer, because an iterative process is required to solve the equations predicting operating temperatures. The output of the new work could be in tables or charts. In effect these charts would provide information on operating variables which correspond to the chosen sludge influent conditions. Alternatively, the program could be made available to allow designers to vary parameters as necessary to suit the actual condition.

AERATION AND MIXING

The Jewell et al. digester did not achieve complete mixing. Despite modifications to the digester shape, dissolved oxygen concentrations remained stratified, indicating inadequate mixing. Therefore, we have recommended that a submerged turbine aerator could be effectively used to eliminate these problems as well as removing the cause of a deflocculated, poor dewatering sludge. Testing is suggested.

DIGESTERS IN SERIES

As Jewell et al. notes, completely mixed digesters in series may be more desirable than a single completely mixed digester. Theoretically, series

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units ought to achieve a given reduction in a smaller total volume. But a comparison between their single stage and dual-stage digesters by Matsch and Drnevich (1977) did not show the dual stage digester to be any more effective in reducing Total Volatile Solids (TVS).

More work needs to be done in studying series digestion. The variation of reaction rate coefficients between basins needs to be examined. Also, the economics need to be studied. Digesters in series appear to be economically attractive because of the theoretical volume reductions. However, these savings may be offset with the costs associated with multiple digester structures. The economic attractiveness of digesters in series may be limited to situations where a more complete reaction is required or where it is desirable to control the process in stages.

HEAT EXCHANGERS

Heat exchangers have been suggested to heat incoming sludge with excess heat from the effluent sludge. Preheating would allow for thinner sludges to become thermophilic and would decrease required detention for a properly insulated container operated with a high efficiency aerator. However, the data and simulation models presently available do not allow for predicting the results. Work is needed in this area, in order that the thermodynamics and economics of exchanging heat in ATAD can be evaluated.

ATAD WITH OXYGEN

The use of pure oxygen to aerate an ATAD system may be economically attractive at a facility where pure oxygen is readily available and being used for other processes such as oxygen activated sludge. Andrews and Kambhu (1970) show that the retention time for 40 percent TVS destruction can be decreased from 7.5 days to 5.5 days. This represents a 25 percent reduction in digester volume when pure oxygen is used as the aerating gas.

SCALE UP

Most of the operating parameters of ATAD can be estimated, using good engineering judgment and the work of Andrews and Kambhu (1970). Nevertheless scale-up is desireable before very large units are constructed. These demonstration tests could be constructed at plants of about 3,800 m^3/d (1 mgd) or less.

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SECTION 4

TEST FACILITIES

The study was conducted primarily at the Binghamton-Johnson City Sewage Treatment Plant located in Binghamton, New York. Present flows average about 68,000 m^3/d (18 mgd). The sludge tested by Jewell et al was a gravity-thickened blend of primary and waste activated sludge. The thickened sludge fell within the range of 1.5 to 7.0 percent total solids with an average of 4.5 percent.

Three scales of testing equipment were used: batch, bench-scale, and full-scale.

BATCH-SCALE EQUIPMENT

The batch reactors consisted of 5.3-l and 6.5-l cylindrical-plexiglass containers placed in water baths. Only 3 liters of sludge were placed in the reactors to facilitate the control of foam. These reactors were used to estimate the biodegradability of sludge.

BENCH-SCALE EQUIPMENT

These units were plexiglass cylinders having a diameter of 25 cm and a height of 61 cm. These units were filled about half to two-thirds with sludge, and aerated with diffused air. During this time, they were held in a water bath to maintain a temperature of 50°C. (A bath was needed because the heat losses from the small apparatus were too great to allow high temperatures to be reached). Foam in these tests was controlled by installing a laboratory stirrer between 5 and 20 cm above the liquid level. Placement of the stirrer depended on the amount of foam produced.

One of these reactors was maintained in a semi-continuous mode, while the other was operated in a continuous mode. In the continuous reactor, sludge was added once per hour by means of a pump. This reactor was use to study systems with a hydraulic retention time of about three days. The semi-continuous reactor received sludge which was added manually twice daily. It was used to study systems having a hydraulic retention of about 5 to 10 days.

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Both reactors were used for long-term testing in conjuction with the full-scale equipment. They received strained sludge at 10°C.

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FULL-SCALE EQUIPMENT

Full scale equipment was utilized specifically for the following: to examine the oxygen transfer efficiencies of the two commercial aerators, to provide a source of information used by the authors in formulating a heat-balance relationship, to examine the kinetics of the process, and finally to test for pathogen destruction.

Before testing, sludge was stored in an insulated storage tank. This tank was mixed by a turbine mixer time controlled to operate for fifteen minutes every two hours.

A centrifugal pump delivered sludge from the storage tank to the reactor. The pump was timer controlled to operate for a number of minutes each hour, thereby allowing control of the retention time.

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The 28-m^3 (1,000-cu ft) steel, circular reactor was totally enclosed and insulated with 10 cm of polyurethane foam. To provide the proper geometry for mixing, the liquid depth in the reactor was set equal to one-half of the reactor diameter, thereby yielding an effective depth of 2.13 meters. To further aid in mixing, the intersection between the floor and wall of the reactor was filled with concrete. A freeboard for foam dissipation was provided by an additional 1.5 meters of reactor height above the operating liquid depth.

During most of the study, the Delaval Separator Company "Centri-rator" aerator was used. This aerator was suspended in the liquid by means of four styrofoam blocks set to insure that the optimum immersion depth was maintained for the impeller.

The other aerator studied was an LFE Corporation "Midland-Frings" aerator. This aerator sat on a tripod 30 cm. from the reactor bottom with the impeller (turbine) spinning within stator plates to draw air down the intake tube and move the liquid and air mixture to the reactor perimeter. The Midland-Frings impeller was encased within the stator, whereas the Delaval Centri-rator impeller hung free in the liquid. Figure 1 shows the operating positions of these aerators within the digester.

A mechanical "foam cutter" was used to control the foam created biologically and chemically. The unit was equipped with a bladed circular impeller and mounted near the outer edge of the top of the reactor. The impeller operated 1 meter above the liquid surface.

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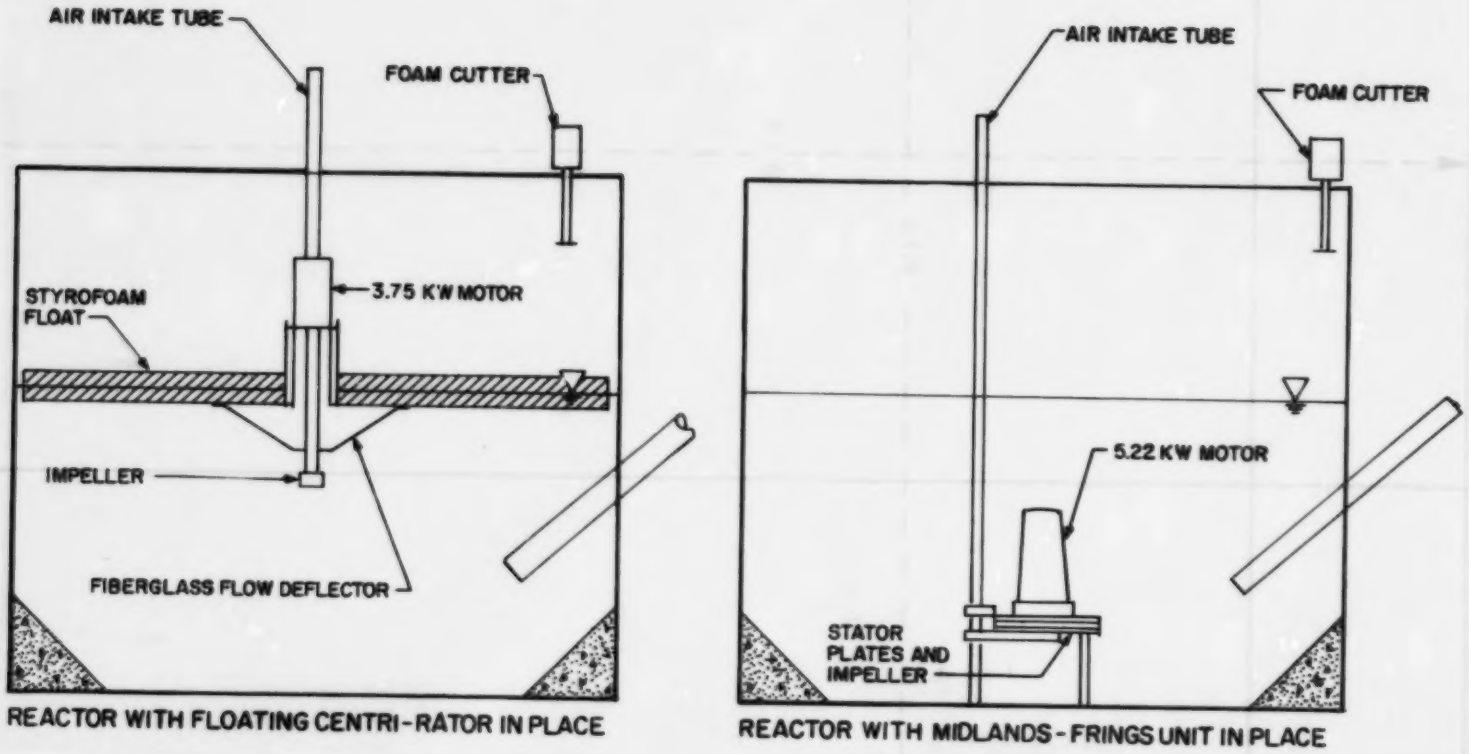

FIGURE I COMPARISON OF THE TWO AERATORS TESTED IN THIS STUDY. AND THEIR LOCATION IN THE REACTOR

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The effluent storage tank was identical to the influent storage tank. It was set approximately 1.5 m into the ground to facilitate gravity overflow from the reactor. This tank enabled collection of samples that were representative of the entire test period and thus minimized variations during the test period. The effluent tank contents were mixed with the same self-priming centrifugal pump that was later used to evacuate the tank.

OPERATIONAL PROBLEMS IN FULL-SCALE TESTING

Throughout the eighteen months of operation, Jewell et al experienced relatively few operational problems. Weather-related problems were heavily influenced by the severity of that year's weather and the temporary nature of an experimental system. Acclimation of the microbial population did not seem to be a problem. The motorized equipment experienced only minor mechanical problem.

The Centri-rator aerator was not affected by rags or similar debris. However, a sludge layer accumulated on the float system and decreased its bouyancy. Consequently, the immersion depth of the impeller increased. This problem could be solved by periodic cleaning.

The Midland-Frings aerator was subjected to rags and debris, which were trapped by the stator plates. The build-up inhibited the rotation of the impeller and caused the overload protection to stop the motor. During hot and humid weather, there were complaints against the stack gases which came from the digester. The odor was similar to a properly operating activated sludge unit. It was found to be objectionable to some people.

The Centri-rator and the Midland-Frings aerators did not achieve complete mixing in the reactor. Dissolved oxygen profiles (Figure 2) showed that dissolved oxygen levels were distinctly stratified. Near the surface, dissolved oxygen ranged from about 2 to 6 mg/l. Near the bottom, dissolved oxygen was less than 0.3 mg/l. The failure to achieve complete mixing makes comparisions with theoretical models somewhat questionable, since these models are based on completely mixed systems.

STEADY-STATE TESTING

Jewell et al operated the full-scale and bench-scale systems in a given mode until steady-state conditions were achieved. To accomplish the steady state, they operated the systems for periods of at least three hydraulic retention times after changing major parameters.

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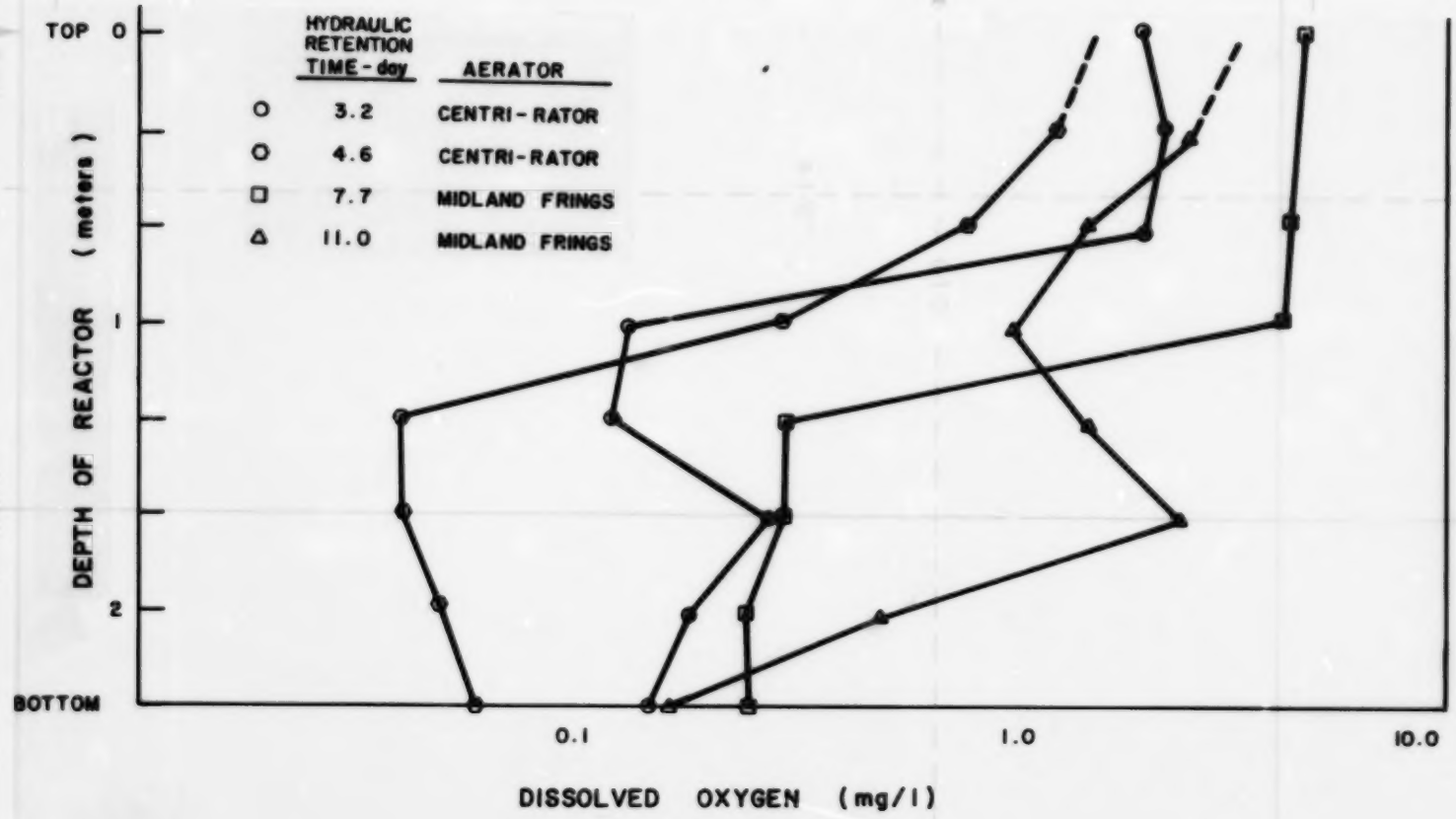
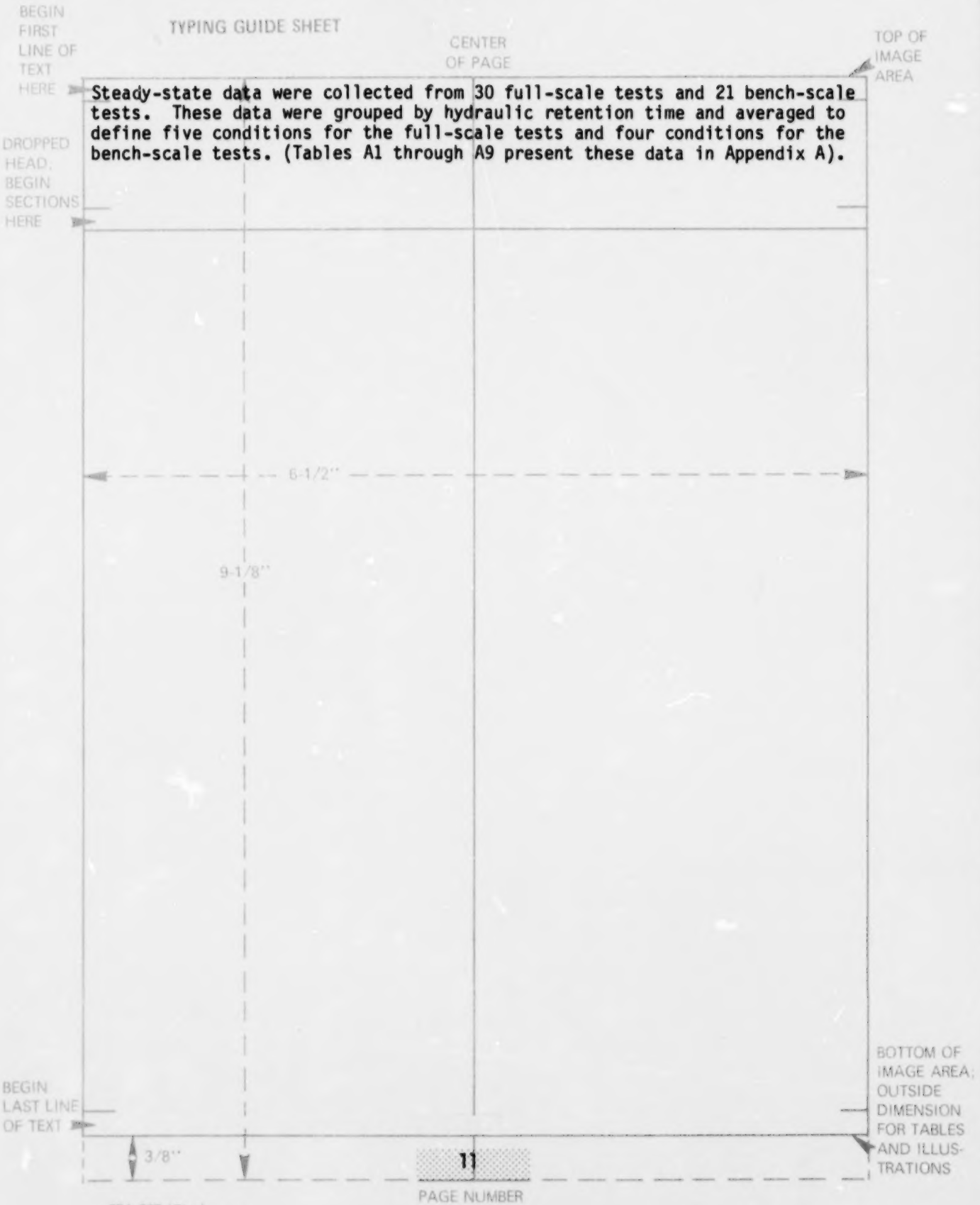


FIGURE 2 DISTRIBUTION FOR DISSOLVED OXYGEN CONCENTRATIONS WITH DEPTH WITHIN THE DIGESTER (FIGURE 65 JEWELL ET AL.)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Steady-state data were collected from 30 full-scale tests and 21 bench-scale tests. These data were grouped by hydraulic retention time and averaged to define five conditions for the full-scale tests and four conditions for the bench-scale tests. (Tables A1 through A9 present these data in Appendix A).
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SECTION 5

TEST RESULTS

In this section, we examine and interpret the test results. For convenience, this chapter is broken into the following topics: Kinetics, Heat Balance, Aerator Transfer Efficiency, Pathogen Destruction, Dewaterability, and Heavy Metal Interactions.

KINETICS

Jewell et al examined the results from their systems using kinetics for completely mixed systems. Because complete mixing was not achieved in their tests, kinetics for completely mixed systems do not totally model what happened under the test conditions. Nevertheless, the use of such a model does allow for some interpretation of the test results.

Using first-order kinetics, the rate of change in biodegradable organics is dependent on concentration of biodegradable organics as shown by:

$$R_S = - K S \qquad (1)$$

where: R_S = rate of change of components
S = concentration of biodegradable organics
K = reaction rate coefficient

The true rate equation is more complex than this first-order equation. However, this approximation has been acceptable in sanitary engineering practice and appears to be reasonable for aerobic thermophilic digestion.

Biodegradability

In equation (1), the "S" is the biodegradable fraction of the sludges at a particular point in time. Because commonly used sludge parameters (e.g. chemical oxygen demand and volatile solids) include non-biodegradable components, Jewell et al estimated the biodegradable fraction of these parameters in the bench-scale testing.

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In these tests, the reactors received a batch of sludge at the beginning of the test period. The sludge was then aerated for 25 to 31 days and tested once per week. The test results were plotted to determine the non-biodegradable components after long periods of aeration. The components examined were COD, total solids, total volatile solids, and total Kjeldahl nitrogen. The tests results showed that the biodegradable portion of COD ranged from 46.4 percent to 76.5 percent. For total solids, the range in biodegradability was from 22.9 percent to 62.0 percent. For total volatile solids, the range in biodegradability was from 42.2 percent to 71.7 percent. For total Kjeldahl nitrogen the range in biodegradability was from 57.0 percent to 85.0 percent. (Andrews and Kambhu (1970) had found biodegradable fractions between 0.5 and 0.8 for volatile solids.) Note that the fractions can vary by over 60 percent. This range can cause some uncertainty in design.

Complete Mixed Systems

For a completely mixed steady-state system, first-order kinetics also predict a decrease in biodegradable organics dependent on the hydraulic retention time and on the rate of reaction as follows:

$$\frac{S_E}{S_I} = \frac{1}{1 + K\,(HRT)} \qquad (2)$$

where: S_E = biodegradable organics out
S_I = biodegradable organic in
HRT = hydraulic retention time

From this equation the reaction constant K can be determined for the ideal system by measuring only three components: the biodegradable organics into the system, the biodegradable organics out, and the hydraulic retention time. Jewell et al obtained results of between 0.15 and 0.80 per day for the reaction rate coefficient, K.

The reaction rate coefficient can be correlated with operating temperature. As shown in Figure 3, trends produced by EPA and Andrews and Kambhu (1970) show that the reaction rate increases with temperature, although Andrews and Kambhu show that the reaction rate begins to decrease at a temperature of about 55°C. The data from Jewell et al generally fall in the vicinity of data reported previously, except for one point from the Midland-Frings aerator. This value is so far out of line with previous values that it might be in error. The error might have been caused by a difference in biodegradability between the continuous flow in the reactor and the sample in the biodegradable test.

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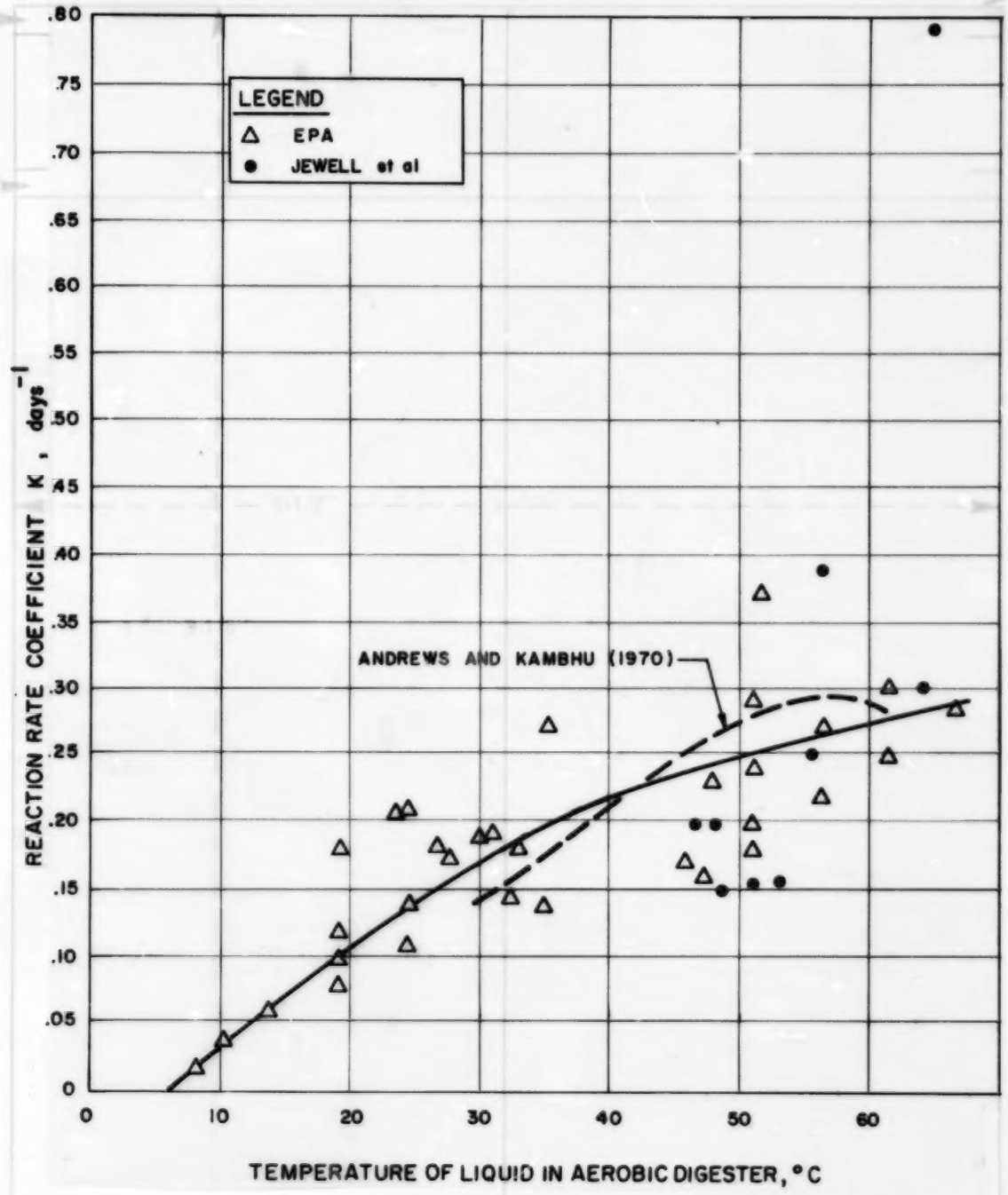

FIGURE 3 REACTION RATE COEFFICIENT VERSUS AEROBIC DIGESTER LIQUID TEMPERATURES

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HEAT BALANCE

Design of ATAD requires that a heat balance around the system be developed, in order to calculate the operating temperature. The heat gains and losses within ATAD can be separated in the following manner:

Heat Gains	Heat Losses
Influent sludge	
Influent gas	Effluent sludge (H_E)
Mixer Input (H_M)	Exhaust gas: Latent heat of vaporization (H_V) Sensible heat loss (H_S).
Biological Production (H_B)	
	Heat loss to surroundings (H_A).

Figure 4 shows these gains and losses schematically.

Jewell et al attempted to simplify the heat balance analysis. First, they neglected the heat in the influent sludge and influent gas. This is reasonable if the reference temperature is close to the temperatures of the sludge and of the gas. Second, they lumped together under a heat loss term (H_L) the sum of the latent heat in the water vapor (H_V), the sensible heat in the effluent gas (H_S), and the heat loss to the surroundings by convection through the surface area (H_A) minus the heat input associated with mixing (H_M). (The terms for heat loss to the effluent (H_E) and for heat production by biological action (H_B) were dealt with separately.) The sum is:

$$H_L = H_V + H_S + H_A - H_M$$

They fit their data with the equation:

$$H_L = a\,(T_R - T_A)^b\, D_L\, C_{PL}$$

where:
T_R = reactor temperature, °C
T_A = ambient air temperature, °C
D_L = liquid density, gm/ml
C_{PL} = specific heat capacity, cal/gm -°C
a,b = coefficients

This empirical relationship was devised in order to eliminate certain system measurements and thus simplify the heat balance.

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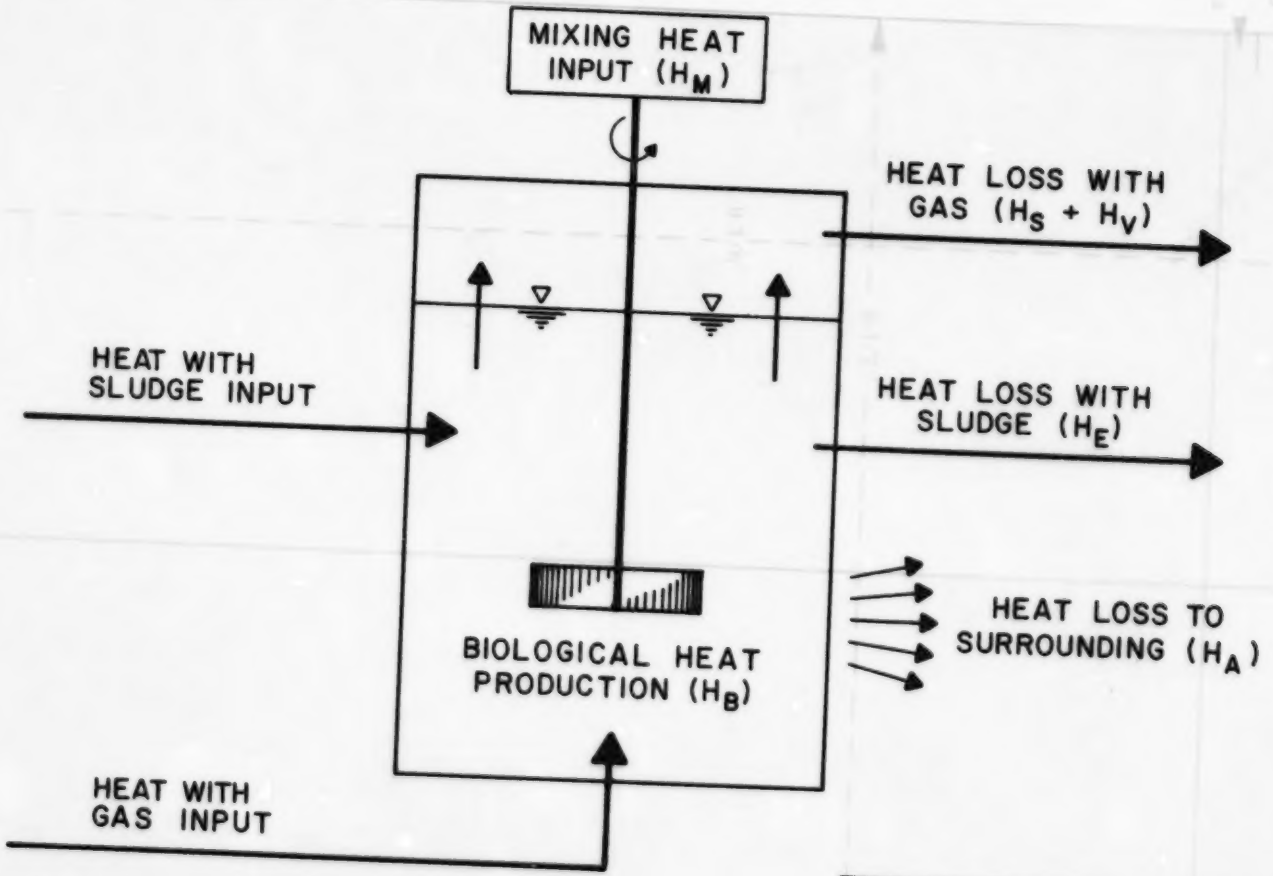
MIXING HEAT INPUT (H_M)
HEAT LOSS WITH GAS ($H_S + H_V$)
HEAT WITH SLUDGE INPUT
HEAT LOSS WITH SLUDGE (H_E)
HEAT LOSS TO SURROUNDING (H_A)
BIOLOGICAL HEAT PRODUCTION (H_B)
HEAT WITH GAS INPUT
$H_B + H_M = H_E + H_S + H_V + H_A$

FIGURE 4 HEAT BALANCE COMPONENTS FOR AEROBIC DIGESTION

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The Jewell et al approach might be applicable to their specific situation, but their formulation should not be applied to other reactors. In general, heat losses are not exponentially related to the difference between reactor and ambient air temperatures. Only the heat loss to the surroundings is related to a difference between the reactor temperature and the ambient air temperature. The mixer input is largely independent of either of these. The losses in the exhaust gas via the heat of vaporization and the sensible heat are independent of the ambient air temperature. These losses depend on the total air flow, on the relative humidity of the exhaust gas, and on the exhaust gas temperature, which would generally be assumed to be about equal to the reactor temperature. Therefore, the heat losses are best estimated independently.

Andrews and Kambhu (1970) used a rational mathematical model to investigate heat losses in aerobic digesters. They found that the most important losses are via the effluent sludge and the exhaust air. While significant, convection losses are of lesser importance and would be controlled with insulation and covering.

The major heat losses can be decreased by providing efficient aerators and by thickening the sludge fed to the reactor. An efficient aerator decreases the air flow through the reactor. A reduction in air flow decreases both sensible heat loss and evaporative heat loss. Andrews and Kambhu (1970) suggested that aerators with efficiencies of 15 percent or better be used for ATAD. Such aerators are available.

Thickened sludge decreases the heat loss by decreasing the total quantity of material that must be raised to operating temperature. At the same time the concentrated sludge provides greater potential for release of heat per unit volume. And, for a given reactor, the increased solids provide for increased detention, thus also raising operating temperatures.

AERATOR TRANSFER EFFICIENCY

Aerator transfer efficiency is a ratio of the gas put into solution to the total gas injected. The maximum transfer efficiency observed by Jewell et al was 22 percent at a loading rate of 9 kg BCOD/m^3·day for the Midland-Frings aerator and 19 percent at a loading rate of 13 kg BCOD/m^3·day for the Centri-rator aerator. The transfer rates for the Centri-rator in tap water was about 3 percent. The Midland Frings unit was not tested in tap water; however, manufacturer's representatives state that the model used is capable of 8 to 10 percent transfer efficiency.

Jewell et al attribute the increase in efficiency in the sludge testing to observed and recorded reductions in bubble size. These reductions are caused by the presence of partially biodegraded organics which reduce surface tension and allow the formation of micro-sized bubbles. These fine bubbles have an increased surface area and retention time by which oxygen can be transferred.

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The value of the measurements of aerator efficiency is debatable. In order for aeration tests to be analyzed, the dissolved oxygen conditions have to be uniform throughout a test basin. As noted earlier, these tests did not achieve complete mixing.

These two aerators are not the only ones that are capable of obtaining high transfer efficiencies. The chemical process industry has been using turbine-type aeration equipment for many years in order to obtain oxygen-absorption efficiencies of 15 to 50 percent for air in thick fermentation broths. The submerged turbine can avoid deflocculation (a problem observed with the two self-aspirating aerators) if peripheral speed does not exceed 4.6 m/s (15 fps).

PATHOGEN DESTRUCTION

One goal of ATAD is to decrease the number of pathogens in the effluent sludge over the number destroyed by conventional digestion. To investigate pathogen kill, Jewell et al compared the numbers of pathogens and indicators in the thickened sludge before it entered the digester with samples drawn from the digesters. About 10 liters of sludge were collected from the thickener pump (raw) and from either the drain valve or sample port on the aerobic digester. Before sampling, Jewell et al disconnected the feed pump for a period of 12 to 24 hours. According to the authors, this procedure, which avoided short circuiting, would be used by operators seeking to maximize pathogen kill. In many operations, however, the digester feed would not be stopped before digested sludge is drawn and pathogen kill might be less.

The authors found that the test results for the aerobic digester were generally better than the results with the anaerobic digester at the Binghamton-Johnson City sewage treatment plant.

Results of bacterial tests were in agreement with those of Smith, Young, and Dean (1975). In the tests by Jewell et al, Salmonella decreased to below detectable limits. In Smith, Young, and Dean's tests, Salmonella species fell to below detectable limits for one bench-scale unit and by more than 2.7 orders of magnitude for another.

In Jewell's tests, Pseudomonas aeruginosa fell by at least 79 percent. In some cases, removal was to below the detectable limit. In the tests by Smith, Young, and Dean, Pseudomonas aeruginosa fell by more than 2.7 orders, to below detectable limits.

In the tests by Jewell et al, fecal coliforms decreased by at least 2.7 orders of magnitude, and, in one case, by more than 4.9 orders. Smith, Young, and Dean reported decreases of 2.4 and 3.8 orders.

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Total coliform decreased by at least 2.5 orders, with one test showing a decrease of more than 5.8 orders. (Smith, Young, and Dean did not report on total coliforms.)

Fecal streptococci fell by 1.6 to more than 4.7 orders of magnitude. In a test at 49°C, Smith, Young, and Dean reported an increase in fecal streptococci. In a test at 56°C, they reported a drop of 2.1 orders of magnitude.

Viruses were virtually completely removed. The only exception occurred when the aerobic digester was having aerator problems.

The low numbers in the feed and effluent samples make interpretation of parasite data difficult. Generally, however, removals of parasites were better than for the adjacent anaerobic digester.

DEWATERABILITY

The effect of thermophilically digested sludge on dewaterability deserves more attention. Jewell et al found that dewaterability was adversely affected by thermophilic digestion in the full-scale reactor. The capillary suction time, CST, increased substantially after digestion. This increase was larger at the higher loading rates. They reasoned that high temperatures create bacterial masses which cause the dewatering problem. Contrary to this, the sludge from their long-term batch test dewatered well. Matsch and Drnevich (1977) had found a thermophilically digested primary plus secondary sludge to be dewaterable on a vacuum filter at 20 $kg/m^2 \cdot h$ (4 lb/sq ft · hr) with modest chemical conditioning costs.

The reason for the poor dewaterability characteristics for sludge from the full-scale tests might be the type of mixer used. The high rotational speeds required by the self-aspirating aerator might have caused deflocculation of the sludge. If this occurred, dewatering characteristics would have been adversely affected. Deflocculation would not be expected with a submerged turbine aerator (as used for ATAD with oxygen) because of the lower rotational speeds. In this report, we examined costs of systems using submerged turbine aerators.

HEAVY-METAL INTERACTIONS

The authors briefly examined the fate of heavy metals within ATAD. They found that the majority of the heavy metals were associated with the particulate phase. Follow-up tests showed that heavy metals could be extracted from the sludge by adding acid to decrease the pH to about 2.

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SECTION 6

ECONOMIC VIABILITY

The economic viability of ATAD should be evaluated by comparing its costs with those of comparable methods of sludge stabilization. This section compares ATAD costs with those of ambient aerobic digestion and mesophilic anaerobic digestion. The comparability of these processes is based on their ability to achieve at least at 40% reduction in TVS.

We have used very conservative criteria to test the economic viability of ATAD. We have been conservative in the sense of choosing a thin sludge which causes more heat to be lost to the effluent than a thickened sludge. The logic is that if ATAD can be shown to be economically successful with this bias, then the process would certainly be economically viable under better conditions.

For this analysis, we selected a sludge having a total solids content of 3 percent. (The sludges examined in the steady-state tests contents ranging from 3.4 to 6.75 percent.) The 3 percent concentration was chosen partly to be conservative and partly because Andrews and Kambhu (1970) had already examined this condition.

We looked in detail at the economics for a 3,800-m^3/d (1-mgd) plant. This is a reasonable size to select, because the first facilities using ATAD would most likely be those in this size range. We estimate that the sludge production from such a plant providing secondary treatment would be about 0.9-dry tonne of sludge per day. We also allowed a sludge-production peaking factor of 50 percent.

In addition to the 3,800-m^3/d (1-mgd) plant we have compared costs of these processes at 38,000-m^3/d (10-mgd) and 380,000-m^3/d (100-mgd) facilites. The selection of digester size and shape serves only as an aid to the economic analysis. These selections do not represent a recommended design.

Our costs do not include those associated with treatment plant administration because these costs would be borne by the normal plant operation. Nor do we include costs for sludge handling either before or after the digestion processes, because we assume that these processes would be the same for each system. Table 1 presents the economic criteria that we used in the cost analyses.

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TABLE 1. CRITERIA USED FOR COST ANALYSIS

Capital Cost Base 3,140 (ENR, April 1980)
Capital Cost Amortization 20 years, 7 percent (Equipment)
40 years, 7 percent (Structures)
Salary $10 per hour (including benefits)
Electricity Cost $0.06/kwh

AUTOTHERMAL AEROBIC DIGESTION

Andrews and Kambhu examined via their mathematical model a sludge having 3 percent solids, 70 percent volatile fraction, of which 70 percent was biodegradable. They calculated that 40 percent decrease in volatile solids would require a detention of 7.5 days with an aerator having an efficiency of 15 percent. Under these conditions, the temperature of the reactor would reach about 32°C. Thus, the digester would not be thermophilic, but it would be taking advantage of self heating. And the stabilization achieved would be about the same as would be expected from anaerobic digestion or ambient aerobic digestion.

The sludge produced in the tests by Jewell et al dewatered poorly, possibly because of the shearing caused by the self-aspirating aerators. As noted earlier, it might be better to use submerged turbine aerators to avoid this. These units can have oxygen transfer efficiencies exceeding 30 percent, depending on the operating conditions.

The required hydraulic retention time for a 3,800-m^3/d (1-mgd) flow could be obtained by constructing two cylindrical digesters, each having 170-m^3 (6,000 c.f.) in volume. To minimize the heat loss, the operating depth should be equal to the diameter. With this condition, the operating depth and the diameter would be 6.1 m (20 ft). A 1.5-m (5-ft) freeboard would be added to allow for foaming. (In design, the actual dimensions might change somewhat to optimize the life cycle costs of the system. A shallower depth, for example, would decrease the power costs for the aerators).

With these dimensions, up to 30 horsepower would be needed to mix and aerate each digester. About half of the power would be required to mix the basin contents. The other half would provide the needed oxygen. At the operating temperature, nitrification would not be depressed, and the oxygen requirement would be about 2.0 kg/kg of volatile solids destroyed.

The capital and annual costs for the 3,800-m^3/d (1-mgd) ATAD system is shown on Table 3. The capital costs include the tanks, mixers, air compressors and compressor building, and appurtenances.

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The major operating expenses are for power and labor. Labor requirements for operation and maintenance should be about the same as for ambient aerobic digestion. According to EPA's Sludge Treatment and Disposal Manual, those labor requirements are about 200 manhours per year.

As shown on Table 2, unit costs are about $160 per tonne ($145/T).

TABLE 2. COSTS OF ATAD FOR 3800 m^3/d PLANT.

Capital Costs	
Turbine Aerators	$ 50,000
Blowers and Piping	60,000
Concrete ATAD Tanks	230,000
Blower Building	45,000
Total (rounded)	$385,000
Annual Costs	
Amortized Capital Costs	$ 31,000
Labor	2,000
Maintenance Supplies	500
Power	19,000
Total (rounded)	$ 53,000
Unit Cost (rounded)	$ 160/tonne

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$A_3$38,000-m^3/d (10-mgd) facility would require a total digester volume of 3,400 m^3 (120,000 cf). This could be achieved by using four digesters 10 m (34 ft) in diameter by 12 m (39 ft) in depth (includes 1.5-m freeboard). These associated costs are:

Capital Costs	$1,200,000)	
Annual Operating Costs	200,000)	Unit Costs = $90/tonne
Total Annual Costs	290,000)	

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A 380,000-m^3/d (100-mgd) facility would require a total volume of 34,000-m^3 (1,200,000 cf). Since minimizing the number of digesters becomes more important than choosing a heat conservative shape; five digesters 30 m (100 ft) in diameter by 12 m (40 ft) deep (includes 1.5-m freeboard) have been chosen. The associated costs are:

Capital Costs	$6,200,000)	
Annual Operating Costs	2,000,000)	Unit Costs = $80/tonne
Total Annual Costs	2,500,000)	

AMBIENT AEROBIC DIGESTION

The critical period for design of an ambient aerobic digester is winter, when temperature is low and biological activity is depressed. In cold climates, the operating temperature might be 7°C or colder. As shown on Figure 3, the reaction rate coefficient (K) for this temperature is only about 0.024 per day. However, we have chosen to evaluate these systems in a climate similar to that which Jewell et al did their work.

To obtain 40 percent decrease in total volatile solids with a 70 percent biodegradable fraction requires that about 57 percent of the biodegradable volatile solids is removed. Equation 2 shows that a detention time of 55 days is required at 7°C for this decrease.

For a 3,800-m^3/d (1-mgd) plant, three digesters, each having the dimensions: 17 m (55 ft) by 9 m (30 ft) by 4.5 m (14 ft) (including 0.6-m freeboard), would provide this retention time. Multiple digesters avoid complete system shutdown in case one digester is out of service due to failure or maintenance of mechanical equipment. And in the summer, one or more units could be shut down as the sludge temperature increases.

Capital and operating costs for a 3,800-m^3/d (1-mgd) ambient aerobic digester system are shown on Table 3. The capital costs include those for the tanks, aerator system, and blower building.

The major operating costs are for power and labor. Mixing and aeration by diffused air would require about 2,900 m^3/h (1,700 cfm). Labor requirements as estimated by EPA in the Sludge Treatment and Disposal Manual would be about 200 manhours per year for operating and maintenance.

Total unit costs would be about $260 per tonne ($240/T).

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TABLE 3. COSTS OF AMBIENT AEROBIC DIGESTION FOR 3800 m^3/d PLANT.

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Capital Cost	
Diffused Air System	
Air Blowers and Piping	$230,000
Concrete Digestion Tanks	250,000
Blower Building	70,000
Total (rounded)	$550,000
Annual Costs	
Amortized Capital Costs	$ 46,000
Labor	2,000
Maintenance supplies	500
Power	39,000
Total (rounded)	$ 88,000
Unit Cost (rounded)	$ 260/tonne

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A 38,000-m^3/d (10-mgd) facility requires a total volume of 16,140 m^3 (570,000 cf). Six free standing structures, each having the dimensions 30.5 m (100 ft) by 15.3 m (50 ft) by 6.7 m (22 ft) (includes 0.6-m freeboard) would provide the necessary volume. The associated costs are:

Capital Costs	$2,500,000)	
Annual Operating Costs	460,000)	Unit Costs = $200 per ton
Total Annual Costs	670,000)	

A 380,000-m^3/d (100-mgd) facility would require a total volume of 160,000-m^3 (5,700,000 cf). Twenty structures, each having the dimensions, 52 m (170-ft) by 26 m (85 ft) by 6.7 m (22 ft) (includes 0.6-m freeboard) would provide the necessary volume. The associated costs are:

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Capital Costs	$14,000,000)	
Annual Operating Costs	4,600,000)	Unit Costs = $180/tonne
Total Annual Costs	5,700,000)	

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ANAEROBIC DIGESTION

An anaerobic digester providing equivalent reduction of volatile solids would require a detention time of about 20 days at average sludge flows. This detention time at a 3,800-m^3/d (1-mgd) flow could be accomplished in two units, each having a diameter of 9.1 m (30 ft) and sidewater depth of 4.6 m (15 ft). At peak sludge flow, detention would be about 13 days.

Mesophilic anaerobic digestion requires that a heat exchanger be used to maintain contents of the digesters at about 35°C. The heat exchanger size is based on the incoming sludge temperature and digester radiation heat losses. Digester gas production provides the fuel required to heat the water used in the heat exchanger.

The capital and operating costs for a 3,800-m^3/d (1-mgd) anaerobic digester system are shown on Table 4. The capital costs include those for the digester, the mixing and heating system, and for a building for heat exchangers and pumps.

The 3,800-m^3/d (1-mgd) anaerobic digestion system requires an estimated 800 manhours per year for operations and maintenance (EPA Sludge Manual). The estimated power requirement for mixing and pumping for heat exchange is 65,000 kwh/year. Fuel for the external heat exchanger boiler is supplied by digester gas.

Total unit cost is about $220 per tonne ($200/T) of sludge.

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TABLE 4. COSTS OF ANAEROBIC DIGESTION FOR 3800 m^3/d PLANT.

Capital Cost	
Digester Mixing System Heat Exchanger and Boiler System Recirculation Sludge Pumps	$340,000
Concrete Digestion Tanks	300,000
Heat Exchanger and Pump Building	60,000
Total (rounded)	$700,000
Annual Costs	
Amortized Capital Costs	$ 59,000
Labor	8,000
Maintenance supplies	2,000
Power	4,000
Total (rounded)	$ 73,000
Unit Cost (rounded)	$ 220/tonne

A 38,000-m^3/d (10-mgd) facility requires a total volume of 5,890 m^3 (210,000-cf). This could be achieved by using four digesters 16-m (52 ft) in diameter by 9.1 m (30 ft) deep (includes 1.5-m freeboard). The associated costs are:

Capital Costs	$1,700,000)	
Annual Operating Costs	42,000)	Unit Costs $55/tonne
Total Annual Costs	190,000)	

A 380,000-m^3/d (100-mgd) facility requires a total volume of 59,000 m^3 (2,080,000 cf). This could be achieved by using eight digesters 30.5 m (100 ft) in diameter by 12 m (40 ft) deep (includes 1.5-m freeboard). The associated costs are:

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Capital Costs	$9,400,000)	
Annual Operating Costs	310,000)	Unit Costs $35/tonne
Total Annual Costs	1,100,000)	

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ECONOMIC COMPARISON

Table 5 summarizes the costs associated with ATAD, ambient aerobic digestion, and anaerobic digestion sludge treatment systems at wastewater treatment plant sizes of 3,800, 38,000 and 380,000-m^3/d (1,10 and 100-mgd).

The least costly alternative for a 3,800-m^3/d (1-mgd) system is ATAD, at about $160/tonne followed by $220/tonne for anaerobic digestion, and $260/tonne for ambient aerobic digestion. Table 5 shows also that ATAD's capital cost is significantly less than those of the other two digestion processes, but that annual operating costs for anaerobic digestion is less than both aerobic digestion and ATAD. Operating costs for aerobic digestion are about twice that of ATAD because of greater power usage.

Anaerobic digestion for both 38,000-m^3/d (10-mgd) and 380,000-m^3/d (100-mgd) systems is the least costly alternative. At these sizes annual operating costs are significantly less than the other digestion processes. Power consumption is the primary reason for the unattractiveness of ATAD and aerobic digestion operating costs.

In summary, the ATAD process for a 3,800-m^3/d (1-mgd) plant size is only about 62 percent of the cost of aerobic digestion, and 73 percent of the cost of anaerobic digestion, on a unit cost basis. Clearly, ATAD merits serious consideration in sludge management studies.

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TABLE 5. ECONOMIC COMPARISON OF ATAD, AEROBIC DIGESTION, AND ANAEROBIC DIGESTION TREATMENT SYSTEMS.

Plant Size	Sludge Digestion Process	Capital Cost	Annual Operating Cost	Amortized Annual Cost	Unit Cost Unit Cost (per tonne)
3,800-m^3/d (1-mgd)					
	ATAD	$ 385,000	$ 21,500	$ 53,000	$ 160
	Aerobic Digestion	550,000	42,000	88,000	260
	Anaerobic Digestion	700,000	14,000	73,000	220
38,000-m^3/d (10-mgd)					
	ATAD	1,200,000	200,000	290,000	90
	Aerobic Digestion	2,500,000	910,000	930,000	200
	Anaerobic Digestion	1,700,000	42,000	190,000	55
380,000-m^3/d (100-mgd)					
	ATAD	6,200,000	1,200,000	2,500,000	80
	Aerobic Digestion	14,000,000	9,000,000	10,000,000	180
	Anaerobic Digestion	9,400,000	310,000	1,100,000	35

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APPENDIX

Performance Data Summary

Table A1. Performance data as collected from the full scale aerobic digester at a design HRT of 3 days (Centri-rator aerator).

Table A2. Performance data as collected from the full scale aerobic digester at a design HRT of 3 days (Midland-Frings aerator).

Table A3. Performance data as collected from the full scale aerobic digester at a design HRT of 5 days (Centri-rator aerator).

Table A4. Performance data as collected from the full scale aerobic digester at a design HRT of 7 days (Centri-rator aerator prior to addition of reactor radius).

Table A5. Performance data as collected from the full scale aerobic digester at a design HRT of 10 days (Midland-Frings aerator).

Table A6. Performance data as collected from a continously fed bench scale digester at a design HRT of 3 days.

Table A7. Performance data as collected from a semi-continously fed bench scale digester at a design HRT of 5 days, first run.

Table A8. Performance data as collected from a semi-continously fed bench scale digester at a design HRT of 5 days, second run.

Table A9. Performance data as collected from a semi-continously fed bench scale digester at a design HRT of 7 days.

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TABLE A1 PERFORMANCE DATA AS COLLECTED FROM THE FULL SCALE AEROBIC DIGESTER AT A DESIGN HRT OF 3 DAYS (CENTRI-RATOR AERATOR).

	IDENTIFICATION AND DATES OF STUDY (1978)									
	4/24-27 #1	4/27-30 #2	4/30-5/3 #3	5/3-6 #4	5/6-9 #5	5/9-12 #6*	5/12-15 #7*	5/15-18 #8*	5/18-21 #9*	5/21-24 #10
θ Days	3.03	3.07	3.05	2.73	3.25	2.83	3.11	3.64	3.21	3.27
TCOD Loading Rate**	21.8	21.0	18.9	20.6	18.2	19.8	18.7	14.6	17.7	16.6
TS Loading Rate**	20.7	18.9	17.3	18.9	17.2	19.8	19.3	14.9	17.7	17.2
TVS Loading Rate**	13.5	12.4	11.3	12.2	10.5	12.1	11.8	9.22	10.9	10.4
TCOD Removal Rate**	5.26	2.86	3.11	1.80	1.85	4.57	5.19	3.52	4.82	3.06
TS Removal Rate**	3.53	1.67	1.29	1.49	1.89	3.38	4.23	1.48	1.87	2.07
TVS Removal Rate**	3.87	3.04	2.27	1.85	1.64	2.84	3.78	2.09	2.47	1.98
TCOD Treatment Efficiency (%)	24.2	13.6	16.5	8.70	10.2	23.0	27.7	24.2	27.2	18.4
TS Treatment Efficiency (%)	17.1	8.80	7.50	7.90	10.9	17.1	21.9	10.0	10.6	12.0
TVS Treatment Efficiency (%)	28.7	24.6	20.1	15.1	15.5	23.5	32.0	22.6	22.5	18.9
TKN Feed (mg/ℓ)	- -	- -	- -	- -	2744	- -	2268	- -	- -	- -
TKN Effluent (mg/ℓ)	- -	- -	- -	- -	2107	- -	2023	- -	- -	- -
NH_3-N Feed (mg/ℓ)	- -	- -	- -	- -	256	- -	188	- -	- -	- -
NH_3-N Effluent (mg/ℓ)	- -	- -	- -	- -	485	- -	630	- -	- -	- -
pH Feed	5.6	5.7	5.8	5.9	5.7	5.8	5.7	- -	5.6	5.6
pH Effluent	6.7	7.0	7.1	5.7	6.0	6.8	6.8	- -	6.9	6.7
Temp. Reactor (°C)	54.9	53.6	46.6	36.0	42.0	46.9	50.4	52.9	49.8	45.8
Temp. Ambient (°C)	8.1	9.5	5.2	7.3	10.7	14.1	14.6	12.6	16.5	14.6
Temp. Feed (°C)	9.1	9.6	10.0	9.9	10.8	10.8	11.8	12.1	12.6	13.2
Capillary Suction Time Feed (secs)	2232	1979	1813	1236	830	531	759	332	377	411
Capillary Suction Time Effluent (secs)	1662	2576	1696	1342	1172	1084	1018	1223	1053	812
Airflow (ℓ/sec)	29.8	30.4	27.0	26.0	25.6	25.9	26.0	25.5	26.5	25.9
Oxygen Transfer Efficiency (G.C.)(%)	23.1	- -	22.8	- -	18.2	- -	23.1	22.8	- -	21.9
Oxygen Transfer Rate (G.C.)(kg/kW-hr)	2.05	- -	1.86	- -	1.25	- -	1.98	1.90	- -	1.77
Oxygen Transfer Efficiency (COD Basis)(%)	22.5	12.3	15.1	9.13	9.73	19.5	21.6	14.9	19.4	12.9
Oxygen Transfer Rate (COD Basis)(kg/kW-hr)	1.98	1.00	1.14	0.63	0.64	1.67	1.86	1.24	1.71	1.04
Duration of Test (day)	2.54	2.60	2.98	2.92	3.00	2.83	2.98	2.85	2.92	2.94

*Steady state averaging points.

**kg/m^3-day.

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TABLE A2 PERFORMANCE DATA AS COLLECTED FROM THE FULL SCALE AEROBIC DIGESTER AT A DESIGN HRT OF 3 DAYS (MIDLAND-FRINGS AERATOR).

	IDENTIFICATION AND DATES OF STUDY (1978)		
	7/26-29 #11	7/29-8/1 #12*	7/24-26 #13
θ_H (days)	2.87	3.49	4.66
TCOD Loading Rate**	13.6	14.6	12.5
TS Loading Rate**	13.5	15.6	8.56
TVS Loading Rate**	8.78	9.75	5.77
TCOD Removal Rate**	4.53	6.53	5.14
TS Removal Rate**	4.64	8.30	3.13
TVS Removal Rate**	2.80	4.89	1.98
TCOD Treatment Efficiency (%)	33.0	44.6	44.2
TS Treatment Efficiency (%)	31.9	50.2	36.6
TVS Treatment Efficiency (%)	34.3	53.3	34.4
TKN Feed (mg/l)	- -	- -	- -
TKN Effluent (mg/l)	- -	- -	- -
NH_3-N Feed (mg/l)	- -	- -	- -
NH_3-N Effluent (mg/l)	- -	- -	- -
pH Feed	- -	5.4	- -
pH Effluent	- -	6.80	- -
Temp. Reactor (°C)	62.0	61.8	59.9
Temp. Ambient (°C)	21.0	17.7	23.3
Temp. Feed (°C)	20.5	20.0	21.5
Capillary Suction Time Feed (secs)	241	75	- -
Capillary Suction Time Effluent (secs)	628	553	698
Airflow (g/sec)	39.6	39.6	40.0
Oxygen Transfer Efficiency (COD Basis) (%)	15.0	22.1	17.5
Oxygen Transfer Rate (COD Basis) (kg/m^3-hr)	1.28	1.80	1.46
Duration of Test (days)	2.83	3.06	3.96

*Used for steady state averaging.

**kg/m^3-day.

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TABLE A3 PERFORMANCE DATA AS COLLECTED FROM THE FULL SCALE AEROBIC DIGESTER AT A DESIGN HRT OF 5 DAYS (CENTRI-RATOR AERATOR).

	IDENTIFICATION AND DATES OF STUDY (1978)						
	#14 2/20-25	#13 2/25-3/2	#16 3/2-7	#17* 3/7-11	#18* 3/11-15	#19* 3/15-20	#20* 3/20-26
θ Days	5.71	5.55	4.60	4.67	4.39	4.77	4.41
TCOD Loading Rate**	9.40	9.86	11.1	11.1	11.9	12.0	13.3
TS Loading Rate**	9.75	9.39	10.3	10.8	11.0	12.4	13.9
TVS Loading Rate**	6.08	5.94	6.60	6.93	6.98	7.55	8.44
TCOD Removal Rate**	3.72	2.68	2.77	3.23	3.26	4.32	4.11
TS Removal Rate**	3.25	0.64	0.38	1.99	2.12	3.31	2.88
TVS Removal Rate**	2.62	1.29	1.20	2.05	2.01	2.71	2.64
TCOD Treatment Efficiency (%)	39.6	27.2	25.0	29.1	27.3	36.1	30.8
TS Treatment Efficiency (%)	33.4	6.87	3.70	18.4	19.4	26.7	20.7
TVS Treatment Efficiency (%)	43.0	21.7	18.2	29.5	28.7	35.9	31.2
TKN Feed (mg/ℓ)	- -	- -	- -	- -	- -	- -	- -
TKN Effluent (mg/ℓ)	- -	- -	- -	- -	- -	- -	- -
NH_3-N Feed (mg/ℓ)	- -	- -	- -	- -	- -	- -	- -
NH_3-N Effluent (mg/ℓ)	- -	- -	- -	- -	- -	- -	- -
pH Feed	- -	- -	5.7	5.7	5.8	- -	5.9
pH Effluent	- -	- -	7.1	6.9	6.8	6.9	7.1
Temp. Reactor (°C)	37.6	42.5	46.7	49.0	48.0	46.9	47.9
Temp. Ambient (°C)	-7.1	-6.3	-8.3	-3.4	3.0	0.3	4.0
Temp. Feed (°C)	5.2	5.8	5.3	5.5	6.5	6.4	6.6
Capillary Suction Time Feed (secs)	- -	411	426	257	455	229	690
Capillary Suction Time Effluent (secs)	- -	592	703	1012	866	879	740
Airflow (ℓ/sec)	24.0	25.0	25.0	25.0	25.0	26.0	26.0
Oxygen Transfer Efficiency (G.C.) (%)	- -	- -	- -	- -	- -	- -	- -
Oxygen Transfer Rate (G.C.) (kg/kW-hr)	- -	- -	- -	- -	- -	- -	- -
Oxygen Transfer Efficiency (COD Basis) (%)	19.5	13.5	13.9	16.2	16.4	20.9	19.8
Oxygen Transfer Rate (COD Basis) (kg/kW-hr)	1.19	0.85	0.89	1.02	1.05	1.47	1.32
Duration of Test (day)	4.90	4.79	4.94	4.17	4.43	4.56	3.92

*Steady state averaging.

**kg/m^2-day

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TABLE A4 PERFORMANCE DATA AS COLLECTED FROM THE FULL SCALE AEROBIC DIGESTER AT A DESIGN HRT OF 7 DAYS (CENTRI-RATOR AERATOR PRIOR TO ADDITION OF REACTOR RADIUS).

	IDENTIFICATION AND DATES OF STUDY (1977)					
	#21 7/25-8/1	#22 8/1-8	#23 8/8-15	#24[a] 8/15-22	#25[a] 8/22-29	#26[a] 8/29-9/5
θ Days	7.28	7.14	8.02	7.74	7.20	8.27
TCOD Loading Rate[bb]	3.37	3.94	4.29	5.05	5.30	5.18
TS Loading Rate[bb]	4.10	4.31	4.67	4.41	4.72	4.06
TVS Loading Rate[bb]	2.64	2.82	2.96	2.73	2.92	2.68
TCOD Removal Rate[bb]	0.87	1.67	1.99	2.28	2.37	2.41
TS Removal Rate[bb]	0.00	1.38	1.68	1.09	1.42	0.69
TVS Removal Rate[bb]	0.41	1.16	1.25	0.84	1.11	0.90
TCOD Treatment Efficiency (%)	26.0	42.4	46.5	45.1	44.8	46.5
TS Treatment Efficiency (%)	0.00	32.1	35.9	24.7	30.2	17.0
TVS Treatment Efficiency (%)	15.6	41.1	42.1	30.9	37.9	33.8
TKN Feed (mg/ℓ)	- -	- -	- -	1279	1139	1209
TKN Effluent (mg/ℓ)	- -	- -	- -	968[a]	1070	1420
NH_3-N Feed (mg/ℓ)	- -	- -	- -	103[a]	181	135
NH_3-N Effluent (mg/ℓ)	- -	- -	- -	235[a]	280	417
pH Feed	5.5	5.2	5.4	5.5	- -	5.5
pH Effluent	7.1	7.5	7.6	7.6	8.0	7.5
Temp. Reactor (°C)	46.9	48.7	48.5	47.1	45.2	47.8
Temp. Ambient (°C)	19.0	22.7	22.6	16.6	18.8	21.7
Temp. Feed (°C)	21.5	22.0	22.4	21.1	21.4	22.3
Capillary Suction Time Feed (secs)	- -	- -	- -	- -	- -	- -
Capillary Suction Time Effluent (secs)	- -	- -	- -	- -	- -	- -
Airflow (ℓ/sec)	28.5	27.0	27.4	24.3	25.4	25.2
Oxygen Transfer Efficiency (G.C.) (%)	6.1	10.5	7.8	11.6	7.3	7.1
Oxygen Transfer Rate (G.C.) (kg/hr-hr)	0.31	0.57	0.50	0.74	0.51	0.52
Oxygen Transfer Efficiency (COD Basis) (%)	3.2	9.1	10.6	13.7	13.7	14.0
Oxygen Transfer Rate (COD Basis) (kg/hr-hr)	0.31	0.60	0.79	0.85	0.87	0.88
Duration of Test (day)	7.08	6.88	7.00	6.88	7.00	7.00

[a] Steady state averaging

[bb] kg/m^3-day

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TABLE A5 PERFORMANCE DATA AS COLLECTED FROM THE FULL SCALE AEROBIC DIGESTER AT A DESIGN HRT OF 10 DAYS (MIDLAND-FRINGS AERATOR).

	IDENTIFICATION AND DATES OF STUDY (1978)			
	#27 10/16-26	#28 10/26-11/5	#29* 11/5-15	#30 11/15-25
θ Days	11.0	10.2	11.0	10.2
TCOD Loading Rate**	3.40	3.76	4.36	4.23
TS Loading Rate**	3.32	3.56	3.49	3.50
TVS Loading Rate**	2.41	2.58	2.53	2.60
TCOD Removal Rate**	1.02	1.52	2.51	1.77
TS Removal Rate**	1.10	0.82	1.43	0.93
TVS Removal Rate**	0.93	0.79	1.25	0.86
TCOD Treatment Efficiency (%)	30.1	39.2	57.5	41.9
TS Treatment Efficiency (%)	33.2	23.0	41.1	26.7
TVS Treatment Efficiency (%)	38.8	30.7	49.4	33.2
TKN Feed (mg/ℓ)	- -	- -	1435	- -
TKN Effluent (mg/ℓ)	- -	- -	1323	- -
NH_3-N Feed (mg/ℓ)	- -	- -	105	- -
NH_3-N Effluent (mg/ℓ)	- -	- -	650	- -
pH Feed	5.5	5.4	5.7	5.6
pH Effluent	6.8	6.7	7.1	7.1
Temp. Reactor (°C)	54.6	60.9	62.3	65.0
Temp. Ambient (°C)	7.1	6.4	6.8	2.7
Temp. Feed (°C)	13.6	12.7	11.9	10.7
Capillary Suction Time Feed (secs)	500	1.0	44	- -
Capillary Suction Time Effluent (secs)	573	100	145	- -
Airflow (ℓ/sec)	40.0	40.0	40.0	40.0
Oxygen Transfer Efficiency (G.C.) (%)	- -	- -	- -	- -
Oxygen Transfer Rate (G.C.) (kg/kW-hr)	- -	- -	- -	- -
Oxygen Transfer Efficiency (COD Basis) (%)	3.2	4.8	7.9	5.7
Oxygen Transfer Rate (COD Basis) (kg/kW-hr)	0.29	0.40	0.66	0.45
Duration of Test (day)	9.88	9.95	9.94	10.0

*Steady state averaging

**kg/m^3-day

TABLE A6 PERFORMANCE DATA AS COLLECTED FROM A CONTINUOUSLY-FED BENCH SCALE DIGESTER AT A DESIGN HRT OF 3 DAYS.

	IDENTIFICATION*				
	#1**	#2**	#3**	#4**	#5
θ Days	2.85	2.86	2.86	2.88	2.96
TCOD Loading Rate (kg/m^3-day)	13.1	12.8	13.1	13.8	10.6
TS Loading Rate (kg/m^3-day)	11.9	12.3	13.4	12.9	11.2
TVS Loading Rate (kg/m^3-day)	7.33	7.64	8.53	7.93	7.06
TCOD Removal Rate (kg/m^3-day)	4.12	4.14	5.61	6.65	2.18
TS Removal Rate (kg/m^3-day)	1.07	0.89	1.99	2.19	0.62
TVS Removal Rate (kg/m^3-day)	1.99	1.55	3.33	3.50	1.29
TCOD Treatment Efficiency (%)	31.6	32.3	42.8	48.3	20.6
TS Treatment Efficiency (%)	22.1	17.7	36.2	41.4	13.1
TVS Treatment Efficiency (%)	27.2	20.6	39.1	44.0	18.3
TKN Feed (mg/l)	- -	- -	- -	- -	- -
TKN Effluent (mg/l)	- -	- -	- -	- -	- -
NH_3-N Feed (mg/l)	- -	- -	- -	- -	- -
NH_3-N Effluent (mg/l)	- -	- -	- -	- -	- -
pH Feed	6.5	6.9	5.5	7.0	6.7
pH Effluent	7.1	7.3	6.3	7.1	6.9
Temp. Reactor (°C)	55.0	55.0	55.0	56.0	54.0
Temp. Ambient (°C)	15.0	15.0	15.0	15.0	15.0
Temp. Water Bath (°C)	74.0	63.0	65.0	64.0	62.0
Capillary Suction Time Feed (secs)	- -	- -	- -	- -	- -
Capillary Suction Time Effluent (secs)	- -	- -	- -	- -	- -
Airflow (l/sec)	2.0	2.0	2.0	2.0	2.0
Average Reactor Dissolved Oxygen (mg/l)	0.3	1.1	1.7	1.0	0.1

* Study covered a period from July 8, 1977 to September 3, 1977. The HRT is equal to the duration of the test in all cases.

** Steady state averaging points

TABLE A7 PERFORMANCE DATA AS COLLECTED FROM A SEMI-CONTINUOUSLY FED BENCH SCALE DIGESTER AT A DESIGN HRT OF 5 DAYS, FIRST RUN.

	IDENTIFICATION*				
	#6	#7**	#8**	#9**	#10
÷ Days	5.0	5.0	5.0	5.0	5.0
TCOD Loading Rate (kg/m^3-day)	8.78	10.8	11.8	10.7	9.92
TS Loading Rate (kg/m^3-day)	12.0	9.41	10.8	10.3	9.45
TVS Loading Rate (kg/m^3-day)	6.18	6.00	6.63	6.16	6.26
TCOD Removal Rate (kg/m^3-day)	3.77	4.20	4.41	3.81	3.65
TS Removal Rate (kg/m^3-day)	2.35	1.91	2.85	2.24	2.75
TVS Removal Rate (kg/m^3-day)	2.31	1.61	1.87	1.53	2.36
TCOD Treatment Efficiency (%)	42.9	38.9	37.3	35.6	36.8
TS Treatment Efficiency (%)	51.4	20.3	26.4	21.9	29.1
TVS Treatment Efficiency (%)	40.8	26.7	28.2	24.8	37.7
TKN Feed (mg/ℓ)	- -	- -	- -	- -	- -
TKN Effluent (mg/ℓ)	- -	- -	- -	- -	- -
NH_3-N Feed (mg/ℓ)	- -	- -	- -	- -	- -
NH_3-N Effluent (mg/ℓ)	- -	- -	- -	- -	- -
pH Feed	:.9	5.8	5.8	5.8	5.8
pH Effluent	6.8	6.8	6.7	6.7	6.5
Temp. Reactor (°C)	49.0	42.0	51.0	50.1	50.0
Temp. Ambient (°C)	15.0	15.0	15.0	15.0	15.0
Temperature (°C)	65.0	50.0	70.0	70.0	70.0
Capillary Suction Time Feed (secs)	- -	- -	- -	- -	- -
Capillary Suction Time Effluent (secs)	- -	- -	- -	- -	- -
Airflow (ℓ/sec)	2.0	2.0	2.0	2.0	2.0
Average Reactor Dissolved Oxygen (mg/ℓ)	- -	- -	- -	3.7	2.6

*Study covered a period from October 1, 1977 to December 19, 1977. The HRT is equal to the duration of the test in all cases.

**Steady state averaging.

TABLE A8 PERFORMANCE DATA AS COLLECTED FROM A SEMI-CONTINUOUSLY-FED BENCH SCALE DIGESTER AT A DESIGN HRT OF 5 DAYS, SECOND RUN.

	IDENTIFICATION*					
	#11	#12**	#13**	#14**	#15	#16
θ Days	5.0	5.0	5.0	5.0	5.0	5.0
TCOD Loading Rate (kg/m^3-day)	11.5	11.5	12.4	12.4	12.5	12.6
TS Loading Rate (kg/m^3-day)	12.2	12.2	14.2	14.2	11.7	10.5
TVS Loading Rate (kg/m^3-day)	7.01	7.01	8.29	8.29	6.97	6.64
TCOD Removal Rate (kg/m^3-day)	3.38	3.97	4.74	3.44	3.01	4.55
TS Removal Rate (kg/m^3-day)	2.41	2.39	5.09	3.66	1.91	1.56
TVS Removal Rate (kg/m^3-day)	1.84	1.93	3.46	2.56	1.37	1.39
TCOD Treatment Efficiency (%)	29.5	34.6	38.3	27.8	24.2	36.2
TS Treatment Efficiency (%)	19.8	19.7	36.0	25.9	16.3	14.9
TVS Treatment Efficiency (%)	26.2	27.5	41.7	30.9	19.7	20.9
TKN Feed (mg/ℓ)	- -	- -	- -	- -	3010	- -
TKN Effluent (mg/ℓ)	- -	- -	- -	- -	2156	- -
NH_3-N Feed (mg/ℓ)	- -	- -	- -	- -	42	- -
NH_3-N Effluent (mg/ℓ)	- -	- -	- -	- -	62	- -
pH Feed	6.0	6.0	6.0	6.0	5.7	5.7
pH Effluent	7.6	7.5	7.5	7.9	8.0	8.1
Temp. Reactor (°C)	50.0	51.0	50.7	52.8	51.7	50.5
Temp. Ambient (°C)	15.0	15.0	15.0	15.0	15.0	15.0
Temp. Feed (°C)	69.0	69.0	70.0	69.0	69.0	65.0
Capillary Suction Time Feed (secs)	964	964	964	964	1203	517
Capillary Suction Time Effluent (secs)	1033	1235	2077	4032	2326	1877
Airflow (ℓ/sec)	2.8	2.8	2.8	2.8	2.8	2.8
Average Reactor Dissolved Oxygen (mg/ℓ)	- -	- -	- -	- -	- -	- -

*Study covered a period from January 25, 1978 to May 12, 1978. The HRT is equal to the duration of the test in all cases.

**Used for steady state averaging

TABLE A9 PERFORMANCE DATA AS COLLECTED FROM A SEMI-CONTINUOUSLY FED BENCH SCALE DIGESTER AT A DESIGN HRT OF 7 DAYS.

	IDENTIFICATION*				
	#17	#18	#19	#20**	#21**
@ Days	7.0	7.0	7.0	7.0	7.0
TCOD Loading Rate (kg/m^3-day)	6.49	5.11	5.18	5.32	5.94
TS Loading Rate (kg/m^3-day)	6.50	5.31	4.88	5.44	5.75
TVS Loading Rate (kg/m^3-day)	4.10	3.23	2.82	3.56	3.85
TCOD Removal Rate (kg/m^3-day)	3.05	2.05	2.31	2.66	2.74
TS Removal Rate (kg/m^3-day)	2.83	1.78	1.24	2.07	2.13
TVS Removal Rate (kg/m^3-day)	1.78	1.18	0.84	1.84	1.74
TCOD Treatment Efficiency (%)	47.0	40.0	45.0	50.0	46.0
TS Treatment Efficiency (%)	44.0	33.0	25.0	38.0	37.0
TVS Treatment Efficiency (%)	43.0	37.0	30.0	52.0	45.0
TKN Feed (mg/l)	- -	- -	- -	- -	- -
TKN Effluent (mg/l)	- -	- -	- -	- -	- -
NH_3-N Feed (mg/l)	- -	- -	- -	- -	- -
NH_3-N Effluent (mg/l)	- -	- -	- -	- -	- -
pH Feed	5.9	5.9	6.1	6.1	5.3
pH Effluent	6.4	6.8	6.4	6.8	7.4
Temp. Reactor (°C)	49.0	56.0	55.0	55.0	54.0
Temp. Ambient (°C)	15.0	15.0	15.0	15.0	15.0
Temp. Water Bath (°C)	55.0	65.0	66.0	66.0	65.0
Capillary Suction Time Feed (secs)	- -	- -	- -	- -	- -
Capillary Suction Time Effluent (secs)	- -	- -	- -	- -	- -
Airflow (l/sec)	1.9	1.9	1.9	1.9	1.9
Average Reactor Dissolved Oxygen (mg/l)	3.4	3.6	3.8	3.1	1.7

*Study covered a period from August 11, 1977 to September 19, 1977. The HRT is equal to the duration of the test in all cases.

**Used for steady state averaging.